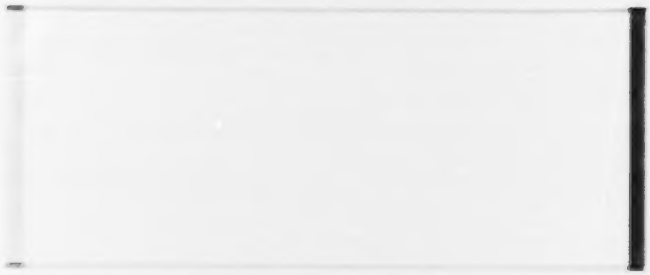

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INVESTIGATION OF SULPHUR CAPTURE DURING CIRCULATING FLUIDIZED BED COMBUSTION OF EASTERN CANADIAN COALS - PHASE I

Final Report

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REPORT NO: MDP/87/04 (revised)
JOB NO: 05000240
DATE: Sept. 2, 1987

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1.0 INTRODUCTION

The Research and Productivity Council was contracted by the Science and Technology Secretariat, Province of New Brunswick, to modify and test RPC's existing circulating fluidized bed test facility. This project constitutes Phase I of a two phase program aimed at investigating sulphur capture during circulating fluidized bed combustion of high sulphur Eastern Canadian Coals. Phase I has been funded by the Science & Technology Secretariat, through the Industrial Innovation and Technology Development subsidiary agreement.

The scope of the work as per the letter of authorization is as follows:

a) modification of stand pipe from 9" to 3" diameter and run three tests, followed by analysis of results,

b) modification of secondary air input to rig, run three tests, followed by analysis of results,

c) review results of (a) and (b) and write report,

d) consultation with the Department of Commerce and Technology representative and Scientific Authority from CANMET.

The project has been completed successfully and all the stated objectives have been met. The system has been modified and operated reliably for over 100 hours of continuous operation. This report describes the work carried out including system modifications, operating experience, experimental procedures, results and discussions.

2.0 DESCRIPTION OF RPC'S FACILITIES BEFORE MODIFICATION

The Research and Productivity Council's circulating fluidized bed pilot-scale test facility is shown schematically in Figure 1. The main components of the system are the combustor, a primary cyclone and a standpipe for solids recirculation, a cooling section, a secondary cyclone and a baghouse.

The Combustor

The cylindrical combustor has a total height of 5.8 m (19') and inside and outside diameters of 12.7 cm (5") and 50.8 cm (20") respectively. The wall has an inner layer of high temperature refractory, a middle layer of castable insulating refractory, and a carbon steel shell. A 0.5 MBtu propane burner is used for start-up purposes.

Primary air is fed into the bottom of the combustor via a distributor plate; a perforated plate with 10% open area located 66 cm (26") from the base of the combustor column. Secondary air can be fed into the combustor via two sets of four diagonal ports at 205.7 cm (81") and 152.0 (6") above the distributor plate.

Stand Pipe

The stand pipe is used for storage and recirculation of solids received from the primary cyclone. It is 2.7 m (9') high and has inside and outside diameters of 22.8 cm (9") and 61 cm (24") respectively. The stand pipe is also refractory lined and is equipped with discharge and sampling ports.

A cooling section is installed at the bottom of the stand pipe. This is used to extract excess heat generated in the combustor so that both bed temperature and air/fuel ratio can be set and maintained at the appropriate levels.

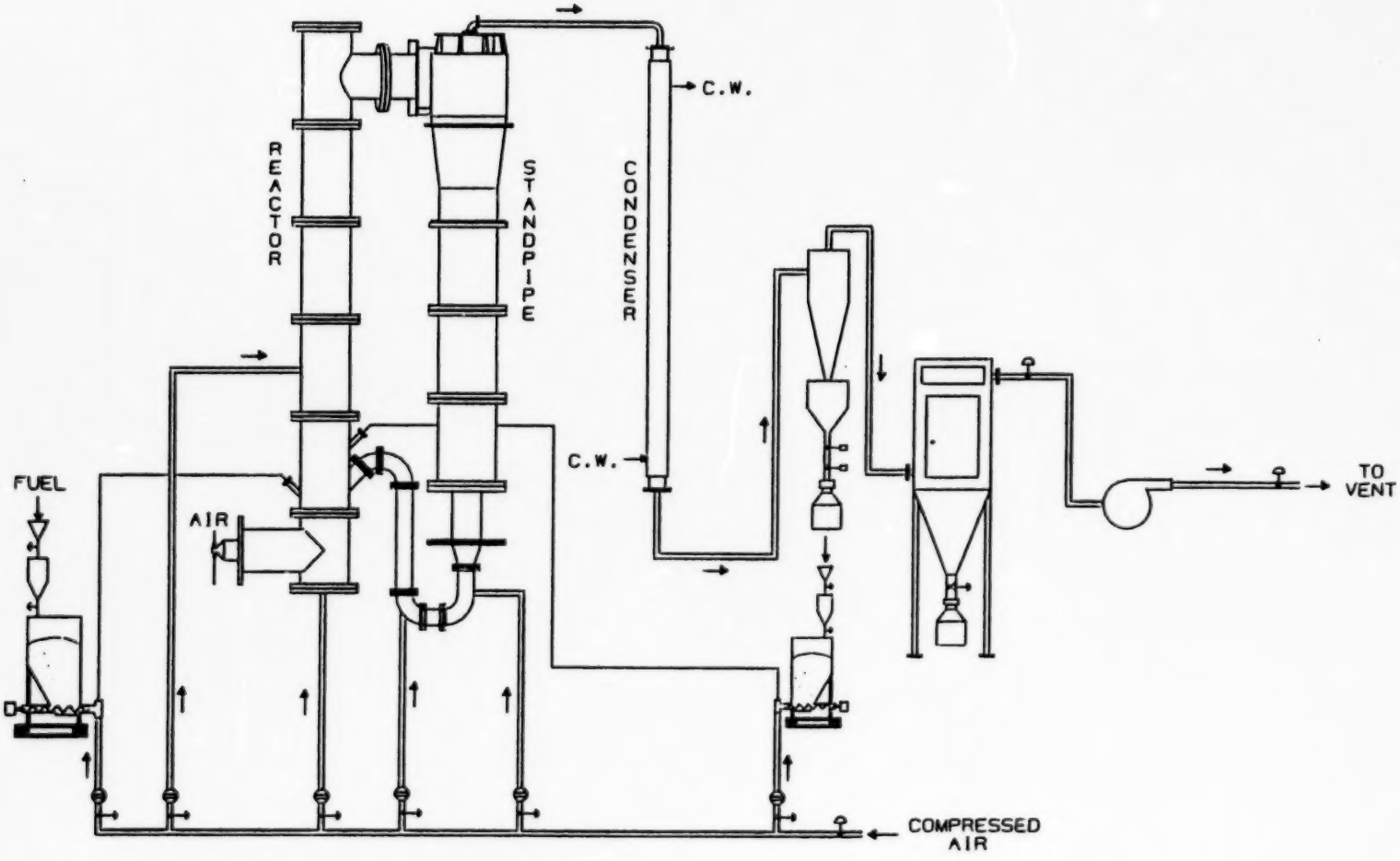

FIGURE 1 : SCHEMATIC OF RPC CIRCULATING FLUIDIZED BED COMBUSTOR

Cyclones

The primary cyclone which is 30 cm (12") in diameter, is refractory lined in a similar manner to the reactor and the standpipe. The secondary cyclone is made of stainless steel and is 25 cm (10") in diameter. The solids collected by the secondary cyclone are recycled pneumatically to the combustor, 48.3 cm (19") above the distributor plate.

J-Valve

Circulation of solids from the standpipe to the reactor is achieved by the non-mechanical J-valve. Air is supplied at two points - one at the base of the standpipe, and the other at the base of the vertical section of the J-valve. The J-valve is made of 2", Schedule 40 stainless steel pipe. The solids from the J-valve enter the combustor at a location 36.8 cm (14.5") above the distributor plate.

Gas Cooling System

Off-gases from the primary cyclone are fed into a water cooled one pass heat exchanger. It is made of two concentric pipes of a nominal diameter of 15 cm (6") (Schedule 40) and 20 cm (8") (Schedule 60), respectively. The height of the exchanger is 5.2 m (17").

Baghouse

The flue gases from the cooling section are fed through the secondary cyclone into the baghouse. The latter is a continuous duty self cleaning dust collector which consists of 16 high temperature bags capable of handling 200 SCFM.

Solid Feeding

Coal/limestone mixtures can be pneumatically fed into the combustor 44.5 cm (17.5") above the distributor plate.

The feed rate is regulated using a screw feeder. The whole assembly is mounted on a weighing scale to provide an accurate record of the feed rate.

Gas Sampling and Analysis System

On-line gas analyzers, (CO, NO_x, SO_2, CO_2, O_2) are available to monitor the composition of flue gases. The gas sampling position is located after the gas cooling system prior to entering the secondary cyclone. The sampled gas is conveyed to the analyzers through a heated line, a gas moisture conditioning system and a diaphragm pump.

Temperature Measurements

Combustor, standpipe and exit gas temperatures are measured with Inconel sheathed K-type thermocouples, located at approximately 30.5 cm (2') intervals along the height of the combustor and the standpipe.

3.0 SYSTEM MODIFICATION AND OPERATING EXPERIENCE

RPC's existing CFBC has been modified in order to reduce the solids capacity of the standpipe. This was achieved by inserting a 3" 316SS pipe inside the 9" I.D. standpipe as shown in Figure 2. The combustor was operated in this new configuration for 60 hours, however some operational problems were experienced as discussed in the first progress report dated June 2, 1987. (see Appendix A)

The main difficulty was due to very low and erratic solid recirculation from the standpipe to the combustor. This led to variation in bed temperature both as a function of time and of location within the combustor. SO_2 capture was low and temperature fluctuations led to clinker formation. The two probable causes for low recirculation rates are loss of head and high resistance to flow within the 3" pipe, located in the standpipe.

As can be seen from Figure 2, any small variation in bed inventory, either due to elutriation or variation in the amount of solid held in the combustor, would lead to large variations in the head of materials in the standpipe. As for resistance to flow within the 3" pipe, some published work suggests that aeration of the standpipe along its height is necsesary for adequate solid flow.*

To overcome the above mentioned difficulties the standpipe was modified again as shown in Figure 3. This configuration has performed very successfully and it's success is probably due to the elimination of any large variation in the head of material and the aeration of the 3" pipe at 2 points, thus improving solid circulation.

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* Hill, M.K., Mallory, R.G. and McKinsey, R.G., "Development of Seal Leg Char Recycle System", Proceding of Ninth International Conference on FBC, Boston, May 1987.

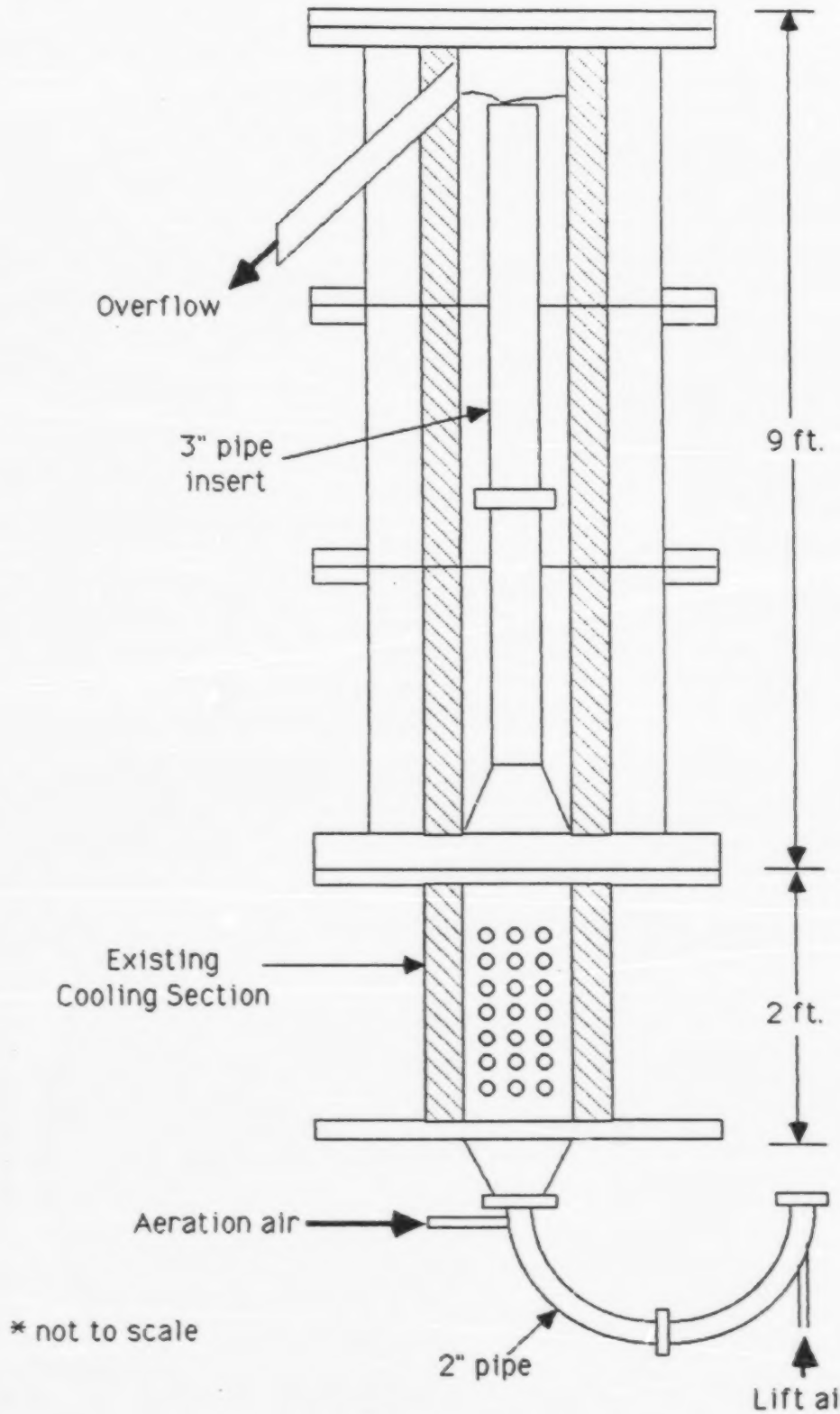

Figure 2: Original Modification of the Standpipe in the CFBC

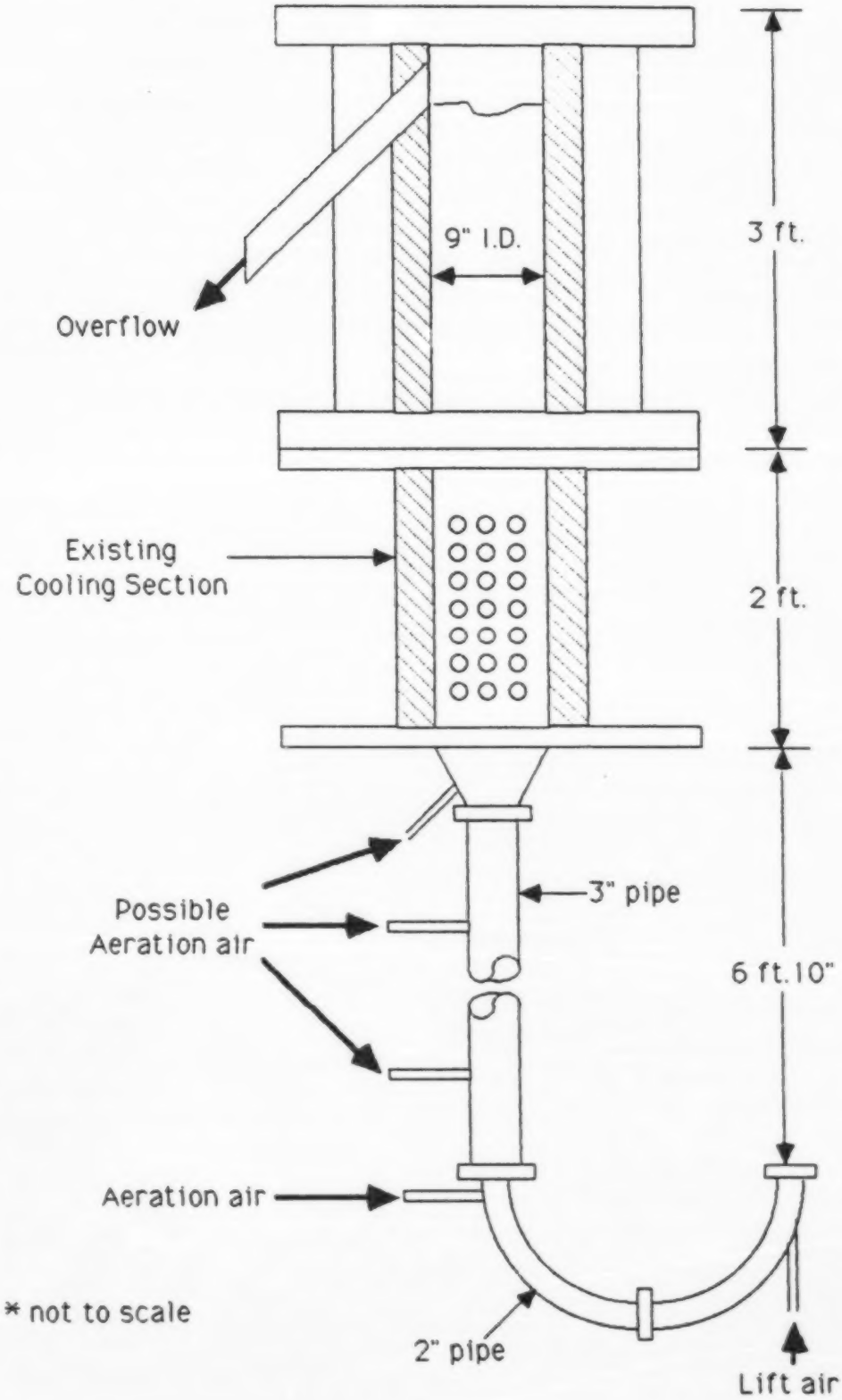

Figure 3: New Modification of the Standpipe in the CFBC

In this mode, the test matrix shown in Table 1 was carried out. The six tests were performed sequentially.

The system was operated successfully in excess of 100 hours with only minor difficulties with the coal feeding system. The most likely explanation for these feeding difficulties is combustor slugging. This condition occurs under certain operating conditions such as high air staging, i.e. lower primary air. To overcome this difficulty the operating pressure at the exit of the primary cyclone was lowered to a value in excess of -15" W.C. This is a fairly high vacuum compared with previous practice where the pressure at that point was balanced to $\pm$5" W.C.

TABLE 1

Test Matrix for the Assessment
of Combustor Modifications

Test No.	Excess Air %	Air Staging secondary/primary	Height of Secondary Air Port Above Distribution Plate, Inch
1	10	40/60	81
2	20	40/60	81
3	20	25/75	81
4	20	25/75	60
5	20	40/60	60
6	10	40/60	60

4.0 EXPERIMENTAL PROCEDURE

The combustor was started on a fresh bed of sand. The 0.5 MBtu/hr propane burner was then switched on and the system was preheated to a temperature of approximately 600°C. At this temperature the coal feed was initiated and the burner was switched off. Once the bed temperature reached 850°C, the cooling section at the bottom of the standpipe was activated and the operating conditions were adjusted for the particular test.

Tests were carried out sequentially, with a transition period between tests which varied from 2 to 6 hours. Test duration was 12 hours. For each test experimental measurements consisted of the following:

(i) temperature measurements along the length of the combustor and standpipe as well as at several other important locations such as the primary cyclone exit, baghouse inlet, etc.

(ii) on-line gas analysis for CO, CO_2, O_2, SO_2 and NO_x.

(iii) all input air flow rates and coal-limestone mixture feed and solid discharge rates from the bed and the baghouse.

During each run solid samples were collected from the baghouse, secondary cyclone, bed discharge and bed material. Samples were collected two, seven and twelve hours after the initiation of the test. Chemical and size analyses were carried out on the samples taken at the seventh hour of each test run. All samples are stored should further analysis be necessary.

5.0 FEEDSTOCK PREPARATION AND CHARACTERIZATION

The Minto coal as received was crushed and screened through 1/8" to remove oversize fractions. The proximate, ultimate and size analyses of the coal are given in Table 2, while Table 3 presents the chemical and size analyses of Elmtree limestone.

TABLE 2
Chemical and Particle Size
Analysis of Minto Coal

PROXIMATE ANALYSIS (wt%)

Moisture	1.0
Ash	15.0
Volatiles	34.1

ULTIMATE ANALYSIS (wt%)

Carbon	69.0
Hydrogen	4.8
Nitrogen	0.9
Sulphur	5.9
Ash	15.0
Moisture	1.0
Oxygen	3.4

CALORIFIC VALUE

BTU/lb	12,750

SIEVE ANALYSIS

+9 mesh	30.3
-9+20	35.0
-20+42	18.0
-42+80	7.2
-170	4.5

TABLE 3
Chemical and Size Analysis
of Elm Tree Limestone

Chemical Analysis (wt%)

Calcium	37.3
CaC03	93.2
Magnesium	0.3
MgC03	0.9

Sieve Analysis

+10 mesh	-
-10 +20	8
-20 +50	40
-50 +100	31
-100	21

6.0 RESULTS AND DISCUSSIONS

Six combustion tests using Minto coal and Elmtree limestone were successfully performed. The tests were carried out consecutively during one week of operation. Each test was 12 hours in duration with conditions set to match the test matrix shown in Table 1. Hourly measurements of feed rates, discharge rates and flue gas composition were recorded and three samples of discharge solids were obtained during each test. The operating conditions are summarized in Table 4. The chemical analyses of the solid samples collected are provided in Table 5. Sample calculations of the mass balances, sulphur reductions, combustion efficiency and excess air are provided in Appendix B.

The particle size distribution of the standpipe bed overflows and samples collected from the J-valve are shown in Table 6. Also shown in Table 6 is the size distribution of the bed emptied from the standpipe at the completion of testing. A comparison of the various size distributions as shown in Figure 4, clearly shows that the bed overflow is not representative of the actual bed while the J-valve samples are representative because of their similarity to the final bed. While the bed overflow was much finer than the bed material no coarsering of the bed is apparent from the cummulative size distribution as shown in Figure 4.

6.1 Mass Balances

The elemental mass balances for C, S, Ca and Ash are shown in Table 7. The ash balance errors obtained for the individual runs were consistently high with an average of 20.1% more ash entering as compared to the recorded ash leaving the system.

TABLE 4
Operating Conditions, Measured
and Calculated Parameters

Parameter	Test Number 1	2	3	4	5	6
Operating Conditions: Feedrates (kg/hr)						
Coal	9.53	8.54	8.77	8.86	9.08	9.46
Limestone	3.73	3.34	3.43	3.47	3.55	3.70
Air	96.33	97.12	99.87	98.98	100.0	99.24
Air Distribution %:						
Primary	32.0	32.0	48.5	47.3	30.5	30.3
Secondary	38.3	38.0	25.5	26.1	40.1	40.1
Others	29.7	30.0	26.0	26.6	29.4	29.6
Measured Parameters:						
Bed Temperature (^{o}C)	846	840	844	834	827	837
Discharge Rate (kg/hr)						
Bed	0.16	1.73	1.03	2.08	1.38	1.86
Baghouse	0.56	0.67	0.92	1.00	0.43	1.06
Cyclone	3.28	1.47	1.12	1.70	2.37	1.98
Cyclone Recycle (kg/hr)	2.07	12.46	14.08	17.82	13.89	17.44
Gas Composition:						
CO_2%	15.07	13.50	14.13	13.83	13.80	14.95
O_2%	3.49	4.85	4.65	4.74	4.87	3.59
CO ppm	200	200	54	229	320	263
SO_2ppm	1617	544	909	641	815	955
NO_x ppm	104	117	132	121	105	81
Calculated Parameters:						
Operating velocity (m/s)	6.7	6.7	6.9	6.8	6.9	6.8
Excess Air %	4.8	17.9	18.1	15.8	15.3	8.8
Sulphur Capture%	69.8	88.6	80.8	86.8	83.2	81.5

TABLE 5

Chemical Analysis of the Solid Samples

Sample	Carbon	Chemical species, wt% Ash	Total Sulphur	Sulphur as Sulphide	Ca	SiO_2
Baghouse						
Run #1	19.4	75.9, 75.7	5.37	1.73, 1.96	10.25	-
#2	17.2	82.1	4.64	1.04, 0.96	11.48	-
#3	12.1	86.3	4.99	0.1	12.1, 11.9	-
#4	13.2	83.0	4.93	0.1	12.2	-
#5	12.4	85.2	5.95	1.13, 1.02	11.3	-
#6	13.4	83.4	5.22	0.4, 0.5	11.9	-
Cyclone						
Run #1	20.0	77.1	4.68	-	19.7	-
#2	11.5	87.5	6.08	-	26.1	-
#3	12.5	89.7	6.67	-	21.7	-
#4	11.9	88.5	5.72	-	21.1	-
#5	13.9	85.4	4.75	-	23.7	-
#6	13.8	85.3	7.18	-	23.4, 24.6	-
J Valve						
Run #1	0.73	97.9	2.19	0.1	8.16	75.5
#2	0.72	98.0	3.79	0.1	14.96	60.2
#3	0.34	96.1	6.43	0.1	23.37	45.0
#4	0.37	98.6	7.28	0.1	24.18	39.8
#5	0.46	98.2	7.68	0.1	27.72	34.9
#6	0.47	98.6	9.14	0.6, 1.0	31.26	27.6
Bed Discharge						
Run #1	14.82	82.6, 82.2	5.30	1.0, 1.31	20.02	28.0
#2	3.66	93.3	5.86	0.2	19.56	40.0
#3	3.49	95.9	7.06	0.1	21.56	35.4
#4	2.50	96.2	6.45	0.1	25.27	31.9
#5	4.20	94.3	8.74	0.6	26.68	24.5
#6	3.30	95.8	8.67	0.1	26.08	28.0

TABLE 6

Particle Size Distribution

of J-Valve Material and Bed Discharge

Sieve Analysis	Run #1		Run #2		Run # 3		Run # 4		Run # 5		Run # 6		Final Bed
	Bed D*	J-Valve**	Bed D	J-Valve	Bed D	J-Valve	Bed D	J-Valve	Bed D	J-Valve	Bed D	J-Valve	
+9 mesh	0.3%	0.4%	0.5%	1.5%	1.2%	0.6%	1.1%	1.4%	0.8%	1.0%	0.8%	1.1%	2.7
-9 + 20	0.5	2.2	1.9	4.6	2.7	4.4	3.8	7.0	2.4	5.0	4.1	5.2	3.4
-20 + 28	-	15.8	10.5	24.8	11.0	20.0	9.8	19.5	10.6	18.7	11.4	14.8	16.2
-28 + 42	12.6	46.4	30.8	48.6	29.0	45.4	27.6	43.2	20.4	42.9	28.9	40.9	35.6
-42 + 80	11.8	13.7	15.2	13.9	15.5	18.0	18.7	16.5	20.1	20.2	20.0	22.5	24.1
-80 + 115	11.3	4.0	8.3	3.8	9.0	6.1	10.2	5.8	11.5	6.1	8.7	7.8	8.7
-115 + 170	11.7	16.6	6.5	1.7	6.3	3.2	7.3	3.4	7.9	3.3	6.2	4.4	4.4
-170 + 270	28.0	0.8	10.5	0.6	10.2	1.9	9.0	1.9	12.5	2.2	10.1	1.9	3.6
-270	23.6	0.2	15.8	0.6	15.1	0.5	12.5	1.3	13.8	0.8	10.0	1.4	1.4

* Bed D = Bed Discharge

** J-Valve = J- Valve Material

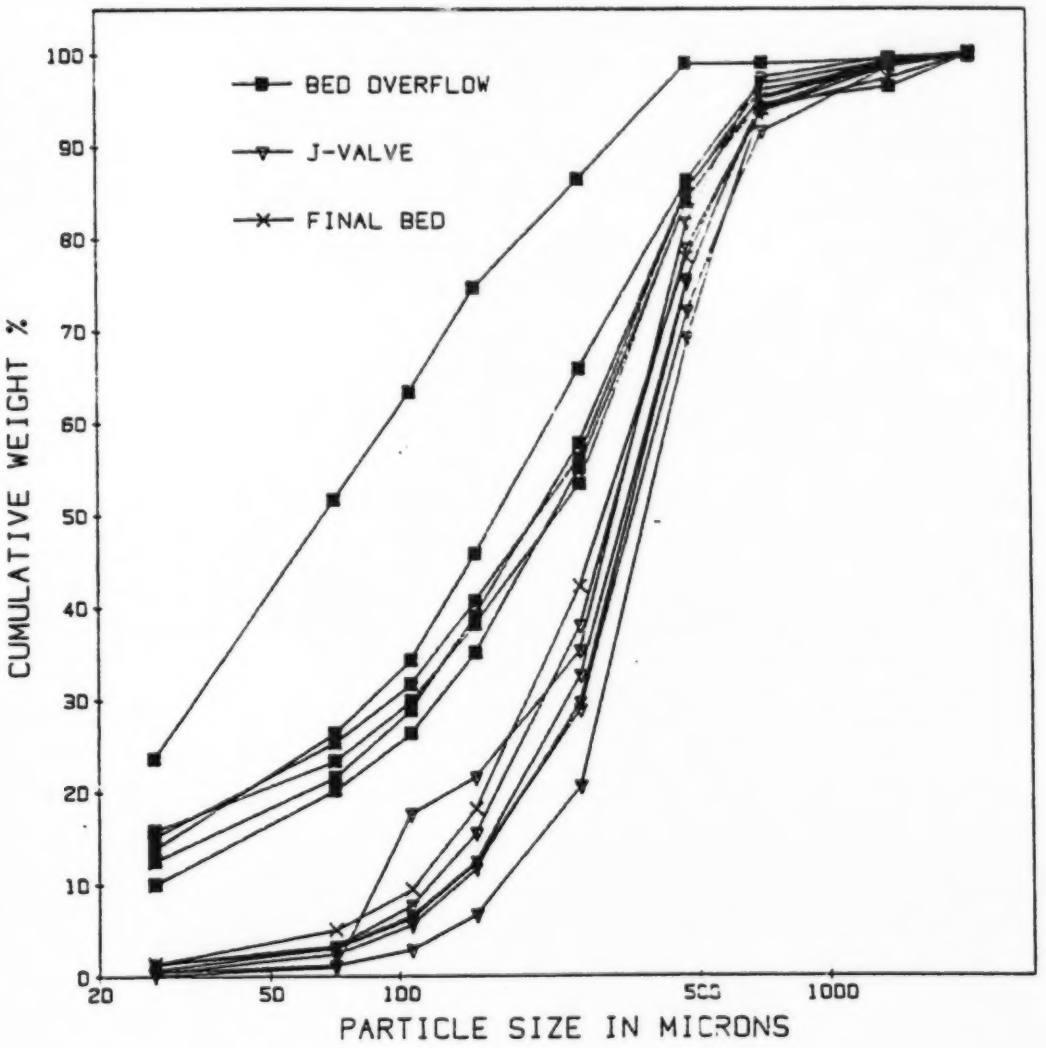

Figure 4: Particle Size Distribution of Bed Overflow, J-Valve and Final Bed Materials

The errors in the ash balances contribute to the errors in the species since all or a portion of the mass reports in the solid streams. As can be seen from a comparison of Table 7 and Table 8, the mass balance errors are decreased by forcing the ash balance to close (ie increasing the total solids discharge from the system). The improvement in the sulphur and calcium balances are significant since all of the calcium is in the solids and the majority of the sulphur is in a solid form (ie $CaSO_4$). The change in the carbon balance is not as significant since the majority of the carbon output is in a gaseous state. A possible cause for insufficent carbon measured in the output stream is the fact that the system was operated under a high vacuum (15" H_2O) and that there was air leakage into the off gas system resulting in a dilution of the flue gas analysis. This would underestimate the carbon content in the flue gas.

The errors in the ash balances are most probably due to the technique used for disposing of secondary cyclone ash. During operation, material collected in the secondary cyclone would be recycled back into the system but on occasions a portion of this ash would be discarded to prevent the recycle rate from growing too large. This would be recorded during a test but if material was discarded between runs it was difficult to allocate a fraction of this material to the total discharge from the preceeding run.

6.2 Combustion Efficiency

The combustion efficiencies for the six tests were determined based on the carbon losses in the ash discharged from the system. Since these values are linked to the accuracy of the measurement of the solids flow rates and there is some doubt as to whether all the ash discharge was properly accounted (see mass balances) two values of combustion efficiences have been provided in Table 9. The first columnn is based on actual measured values and the

TABLE 7

Mass Balance Errors

Run	Carbon	Sulphur	Calcium	Ash
1	4.50	2.58	-1.64	32.9
2	9.86	22.0	4.24	20.8
3	6.54	22.2	32.8	36.3
4	8.73	23.7	12.9	3.67
5	8.48	21.7	5.40	19.1
6	6.88	-5.33	.04	7.89
AVG	7.50	14.5	8.96	20.1

TABLE 8

Mass Balance Errors Corrected for Ash Balance

Run	Carbon	Sulphur	Calcium
1	-1.03	-14.3	-27.5
2	8.41	10.3	-12.6
3	4.00	0.71	6.68
4	8.51	21.6	10.0
5	6.92	10.2	-12.0
6	6.30	-10.9	6.69
AVG	5.52	2.94	-4.79

second column is based on an increased ash discharge ecessary to close the ash balances. As can be seen from Table 9, the combustion efficiences are consistently high even when corrected for the ash balance. The only run with poor combustion efficiency is test 1. The most likely cause for the poor performance of this run is the low amount of material recycled from the secondary cyclone. The carbon content is high in the fine ash collected in the cyclone and not recycling this ash, lowers the residence time of this fraction of carbon considerably since it is quickly elutriated from the reactor.

6.3 NOx Emissions

The NO_x emissions, according to theory, can be decreased by reducing the oxygen available for the oxidation of the fuel bound nitrogen to NO_x. This can be achieved by reducing the excess air supplied and by increasing the fraction of the total air provided by the secondary air (ie. higher air staging). This was investigated in the first part of the test program. In Table 4, the NO_x concentration in the off gas increases from test 1 through test 3, corresponding to an increase in excess air and decrease in the fraction of secondary air.

In the second part of the program the size of the reducing zone in the combustor located below the secondary air ports was reduced by lowering the position of the secondary air ports. It was expected that the NO_x emissions would increase since the residence time of fuel in a reducing atmosphere would be decreased but this was not the case, Table 4. It was found that NO_x emissions were marginally lower with the lower secondary air ports. Tests 4 to 6 show the same trend as tests 1 to 3 ie. lower NO_x emission at higher air staging and lower excess air.

TABLE 9

Combustion Efficiencies
for the Runs

Run	Combustion Efficiency Uncorrected*	Combustion Efficiency Corrected**
1	88.7	83.2
2	94.4	93.0
3	95.5	93.0
4	94.1	93.8
5	93.4	91.8
6	93.2	92.6

* Combustion efficiency calculated on measured solids discharged and carbon content in the discharge.

** Combustion efficiency calculated based on amount of solids discharged corrected to force the ash balance to close.

6.4 Sulphur Capture

The sulphur capture during tests 2 to 6 was consistently above 80% and there appears to be no obvious relationship between sulphur capture and the operating variables tested, (within the ranges tested) namely air staging, excess air and position of the secondary air ports. In test 1 the sulphur capture was low at 69.8%. This can be partially attributed to the low recycle of secondary cyclone dust. This dust, in addition to being high in carbon, has a high CaO concentration. This CaO is very fine and will be highly reactive and does improve the sulphur capture when recycled back into the combustor. Another contributor to the low sulphur capture is the low CaO concentration in the standpipe bed. This material is continually recycled at a high rate and can increase the actual ratio of Ca to S in the combustor appreciably.

6.5 Calcium Sulphide Formation

Previous work carried out at RPC, identified two possible reducing areas within the system where CaS might form. They were the standpipe and the lower section of the combustor. The results given in Table 5 however, tend to suggest that CaS forms at the lower section of the combustor when the system is operated at high air staging. In tests 3 and 4, carried out with low air staging (higher primary air flow), the concentration of sulphide measured in the solids samples was less than 0.1%. By contrast test 1, which had the lowest excess air and the highest air staging, shows the highest sulphide concentration, particularly in the solids reporting in the baghouse.

A comparison of the results from tests 1 to 3 with those from tests 4 to 6 indicate that lowering of the secondary air ports from 81" to 60" did not appear to have an impact on CaS formation. These results should be considered with caution due to the uncertanties associated

with the determination of the forms of sulphur, which has been determined by ASTM method D2492-68. More accuarate analytical methods and further test work will be necessary to quantify the effect of the location of the secondary air ports on CaS formation.

6.6 Time Requirement to Reach Steady State

One of the objectives of this study was to determine the time required for the inventory of solids in the system to come to a steady state composition. Samples were taken from the J-valve throughout the test program and were analyzed for SiO_2, Ca and S. From this analysis it was possible to determine the concentration of various species in the bed as a function of time.

The steady state bed composition will consist of the products of the combustion of coal and limestone. The products are coal and limestone ash, $CaSO_4$ and CaO. The exact composition will depend on the degree of sulphation of the CaO and the degree to which each species is captured by the primary cyclone. For example, the coal ash which is extremely fine may be low in concentration in the steady state bed because of the inefficiency of the primary cyclone in capturing fine particles. A theoretical steady state bed composition assuming equal capture of all species by the primary cyclone and a sulphur capture of 85% is as follows: 34.9% ash, 41.8% $CaSO_4$ and 23.3% CaO. The calculation of this bed composition is shown in Appendix C.

The actual bed composition as a function of time is shown in Table 10. These values are shown plotted in Figure 5. The concentration of SiO_2 in the bed decreases exponentially with time. The expected minimum concentration based on the theoretical bed ash concentration will be approximately 8.9% because of the SiO_2 content of the coal ash. An exponential curve fit of the silica concentration gives the following equation:

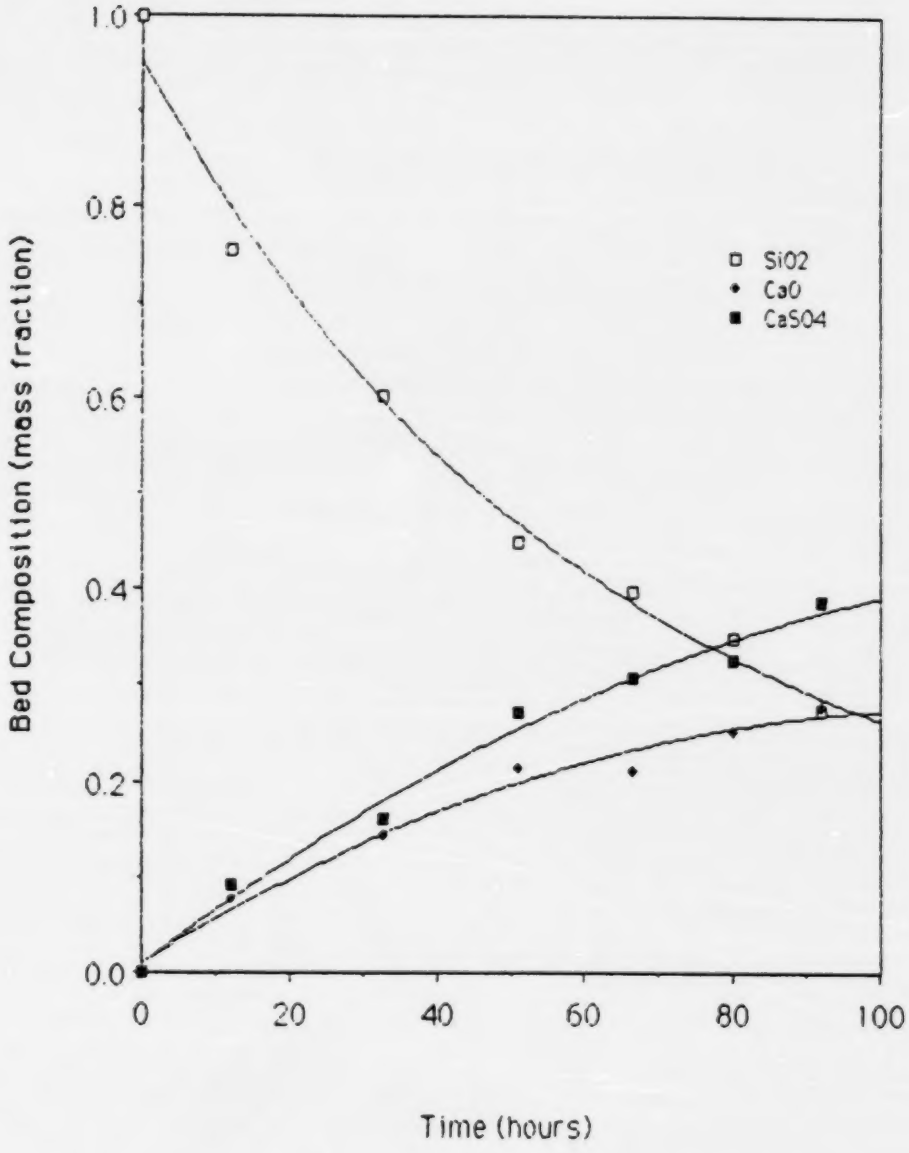

Figure 5: Composition of the Stand Pipe Bed as a Function of Time

TABLE 10

Standpipe Bed Composition

Time (hours)	SiO_2 (wt%)	CaO (wt%)	$CaSO_4$ (wt%)
0	100	0	0
12	75.5	7.59	9.30
32.75	60.2	14.30	16.09
50.9	45.0	21.45	27.30
66.4	39.8	21.10	30.91
79.9	34.9	25.35	32.61
91.9	27.6	27.75	38.81

$$SiO_2\ (t) = .861e^{-0.016t} + .089 \quad (1)$$

where SiO_2 (t) =mass fraction of silica in the bed at time t

t = time in hours

correlation co-efficient = 0.994

The theoretical concentration of SiO_2 in the bed as a function of time based on a simple stirred tank reactor model with an average fuel feedrate of 12.58 kg/hr and corresponding solids discharge rate of 4.57 kg/hr (see Appendix C) is as follows:

$$SiO_2\ (t) = 0.911e^{-.0457t} + 0.089 \quad (2)$$

The mean residence time of a sand particle is equal to 62.5 h based on the curve fit of the actual data while the corresponding theoretical mean residence is equal to 21.9 h. This difference is due to the fact that the system is not a simple stirred tank reactor but can be considered as a stirred tank reactor in series with a solids classifying device, the primary cyclone. The SiO_2 does not break down into smaller particles as quickly as the products from combustion such as coal ash, $CaSO_4$ and CaO and therefore is not elutriated from the system as primary cyclone overflow as quickly as these products. The majority of the SiO_2 must be displaced as bed overflow in the stand pipe, but since this only accounts for 1/3 of the solids discharged from the system it is understandable why it takes 3 times longer to reach a steady level for SiO_2.

The plots of the concentrations of CaO and $CaSO_4$ are more encouraging. The CaO concentration has exceeded the expected steady state concentration after seventy hours of operation and the $CaSO_4$ concentration is close to the expected levels by the end of the tests. Therefore as expected the SiO_2 is the slowest species to reach a steady state equilibrium because of its lower attrition rate compared to other species, but the bed quickly accumulates the important active species CaO which is needed for sulphur capture.

7.0 CONCLUSIONS AND RECOMMENDATIONS

The project has been completed successfully and the main findings are as follows:

1. The system in its present configuration has been operated reliably for over 100 hours of continuous operation.

2. High combustion efficiencies of approximately 93% have been achieved.

3. Although the bed inventory has been reduced to approximately 50% of its original value, the time requirement to remove all the starting sand is still substantial. However CaO and $CaSO_4$ are quickly accumulated in the system and the combustion efficiency and SO_2 capture show little dependence on bed composition.

4. Calcium sulphide formation can be virtually eliminated by appropriate air distribution. In this study only very low (less than 0.1%) concentration of CaS was formed when the secondary air was approximately 25% of the total air flow at an excess air level of approximately 20%.

5. High sulphur capture has been achieved. However, the results show no clear dependance on the operating parameters tested, namely; air staging and excess air levels. These parameters however do have an effect on NO_x formation.

The main recommendation can be summarized as follows:

1. The results of the time required to reach a

steady state bed composition clearly show that in future tests the start up bed should be as free of sand as possible and close to steady state compositions. For this reason it is recommended that the bed material obtained during this Phase should be used in future tests.

2. The effect of parameters such as limestone type and particle size, Ca/S ratio, bed temperature and operating velocity on sulphur capture should be investigated.

AUTHORIZATION

DEPARTMENT: **Mineral Development & Processing**

PROJECT TITLE: **INVESTIGATION OF SULPHUR CAPTURE DURING CIRCULATING FLUIDIZED BED COMBUSTION OF EASTERN CANADIAN COALS (Phase 1)**

PROJECT NUMBER: PROPOSAL NUMBER

REPORT **MDP/87/04**

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APPENDIX A
Progress Report

INVESTIGATION OF SULPHUR CAPTURE DURING CIRCULATING FLUIDISED BED COMBUSTION OF EASTERN CANADIAN COALS - PHASE I

Progress Report No. 1

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REPORT NO: AS/88/04
JOB NO: 05000240
DATE: June 2, 1987

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INTRODUCTION

The main objectives of Phase I as stated in RPC's proposal AS/87/30P, dated April 10, 1987, are to modify RPC's existing test facility and to assess the effect of the modifications on the unit preformance. The scope of modifications includes modifying the standpipe in order to reduce bed inventory and modifying the secondary air ports. A test program is to be carried out after each modification. The progress to date is outlined below.

WORK CARRIED OUT

The progress to date is as follows:

(a) Feedstocks have been acquired, prepared and characterized.
(b) Stand pipe modification has been carried as shown in Figure 1.
(c) The combustor was operating for 60 hours to assess its performance and some operational problems have arisen as discussed below.

The main difficulty was due to very low and erratic solid circulation from the standpipe to the combustor. This has led to variation in bed temperature both as a function of time and of location within the combustor. Also SO_2 capture appears to be low. Temperature swings have led to clinker formation.
The two probable causes for low circulation rates are loss of head and high resistance to flow with the 3" pipe, located in the standpipe.

As can be seen from Figure 1, any small variation in bed inventory either due to elutriation or variation in the amount of solid held in the combustor, would lead to large variations in the head of materials in the standpipe. As for resistance to flow within the 3" pipe, some published work suggests that aeration of the standpipe along its height is necessary for adequate solid flow. To overcome these difficulties the modifications outlined below are recommended.

SUGGESTED FURTHER MODIFICAITONS

The suggested modification shown in Figure 2, reduces bed inventory by ~50% and would eliminate the problems encountered with the earlier design. The large cross-section would overcome the problem of head variations and the smaller section can be aerated to overcome any flow problems.

Figure 1: Original Modification of the Standpipe in the CFBC

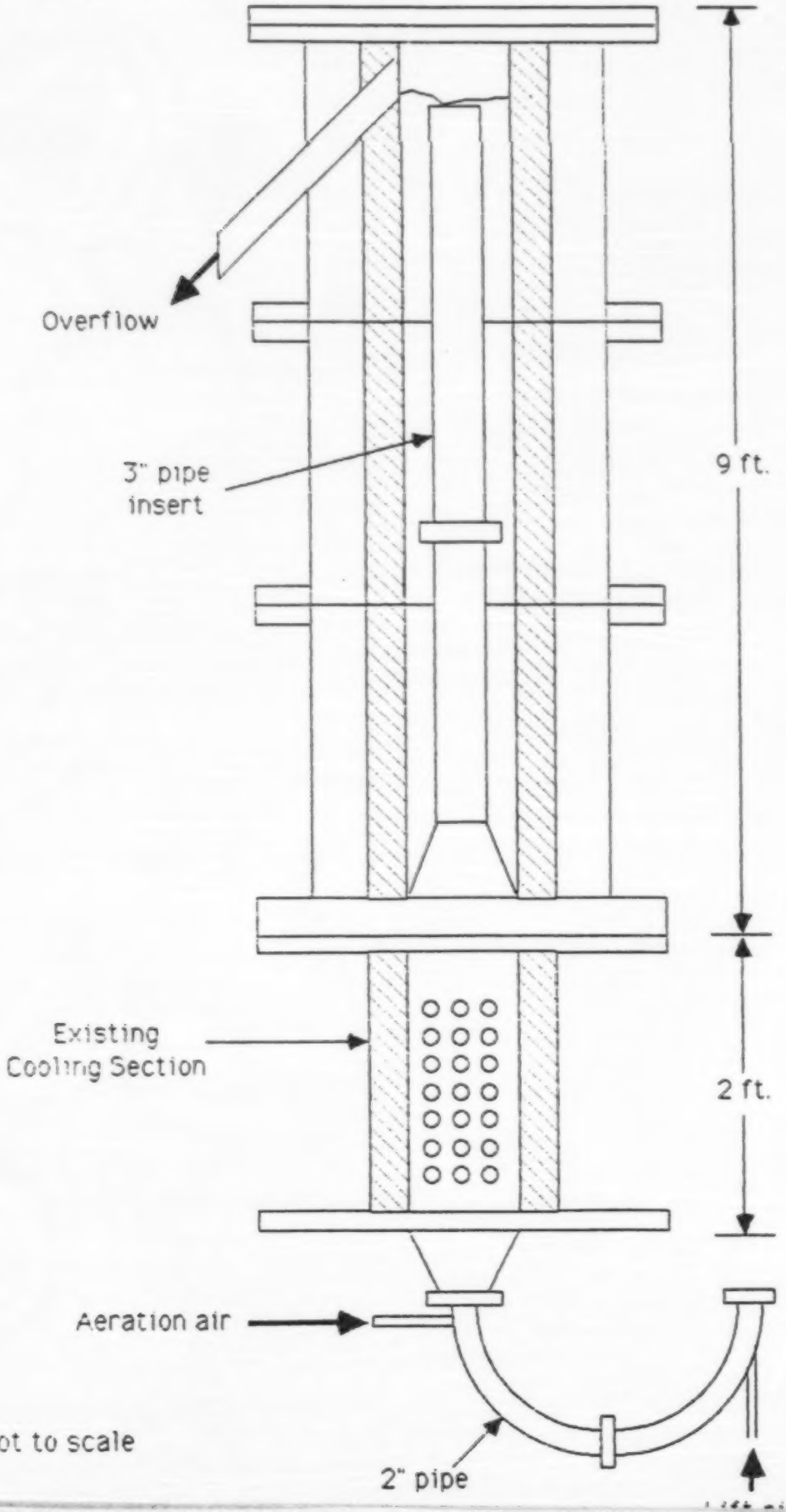
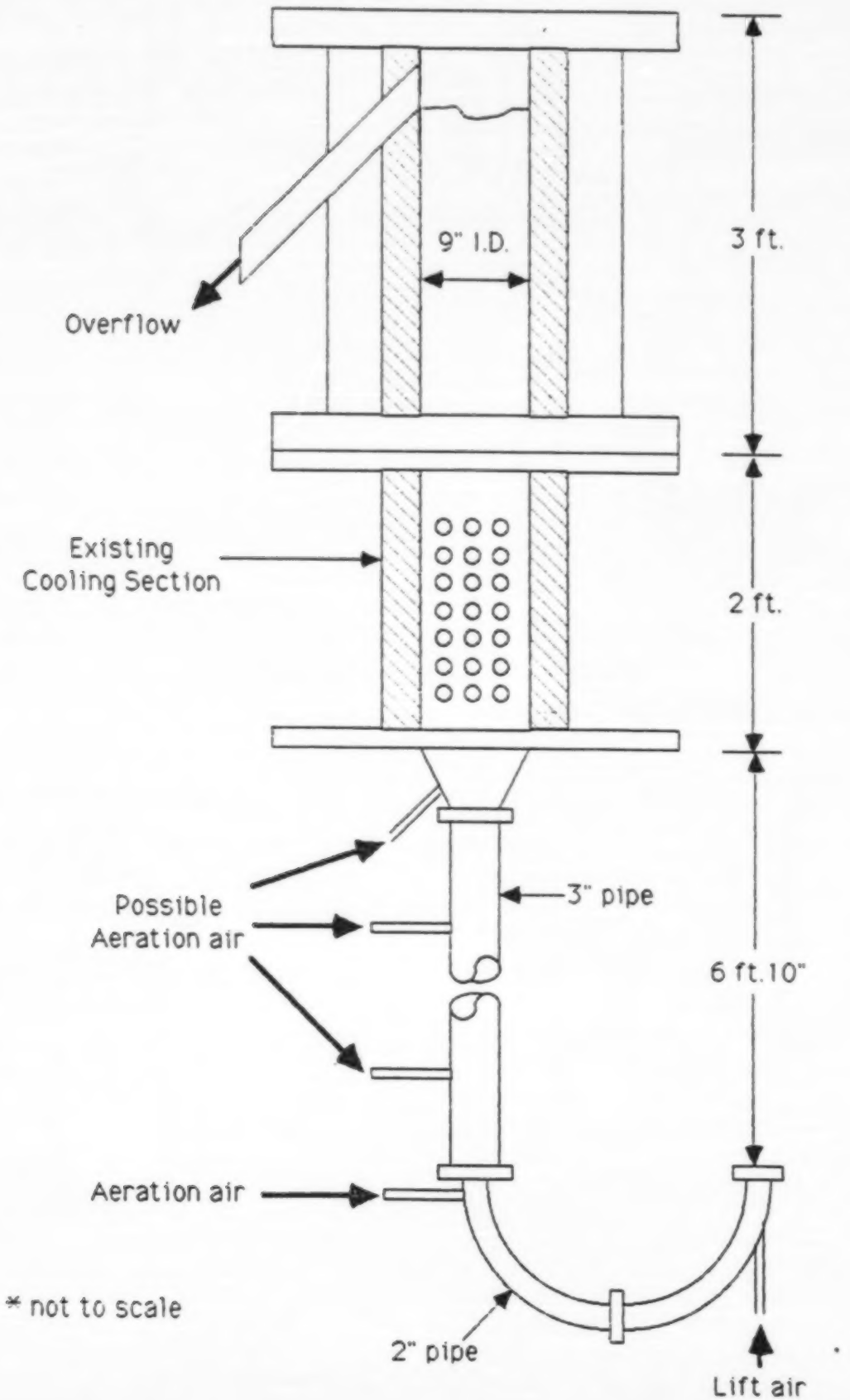


Figure 2: New Modification of the Standpipe in the CFBC



APPENDIX B
Sample Calculations

Sample Calculations

Run 1 basis 1 hour, 100 kg of solids reservoir in the standpipe

Flow rate of coal = 9.53 kg
Flow rate of limestone = 3.73 kg
Flow rate of air = 96.33 kg

Flue gas composition:

CO_2% 15.07
O_2% 3.49
CO ppm 200
SO_2 ppm 1617
NO_x ppm 104

Average molecular weight of the flue gas:
= $\sum$ mole fraction of a species x molecular weight of the species
= 0.1502 x 44.01 + .0349 x 32.00 + .0002 x 28.01 + .001617 x 32.06 + .000104 x 30.00 + .81248 x 28.01

= 30.62 kg/kg-mol

The flowrate of dry flue gas is based on a nitrogen balance

N_2 in:
= mass flow of air x mass fraction N_2 in the air
+ mass flow of coal x mass fraction N_2 in the coal

= 9.53 x 0.0092 + 96.33 x 0.767

= 73.973 kg

= 2.6409 kg-mol

Dry flue gas flow rate = $\frac{\text{kg-mol of } N_2}{\text{mol fraction of } N_2 \text{ in the flue gas}}$

$= \frac{2.6409}{(0.81248 + 0.000104/2)}$

= 3.2501 kg-mol

= 99.52 kg

Carbon in:

= mass flow of coal x mass fraction carbon in the coal
+ mass flow of limestone x mass fraction carbon in the limestone

= 9.53 x 0.6902 + 3.73 x 0.112

= 6.995 kg

Carbon out:

= molar flow of flue gas x mol fration carbon in the flue gas x molecular weight of carbon
+$\sum$ mass flow of solids discharged x mass fraction of carbon in the solids

= 3.2501 x (.1507 + 0.0002) x 12.01
+ (0.16 x 0.1482) + (0.56 x 0.1939) + (3.28 x 0.2005)

= 5.890 + 0.790

=6.680 kg

Carbon balance error:

$= \frac{\text{Carbon in - Carbon out}}{\text{Carbon in}} \times 100\%$

$= \frac{6.995-6.680}{6.995} \times 100\%$

= 4.50%

Sulphur in:

= mass flow of coal x mass fraction S

= 9.53 x 0.0585

= 0.5575 kg

Sulphur out:

= molar flow of flue gas x mol fraction sulphur in the flue gas x molecular weight of sulphur

+ $\sum$ mass flow of solids discharged x mass fraction of sulphur in the solids

+ accumulation of sulphur in the stand pipe bed

= 3.2501 x 0.001617 x 32.06

+ 0.16 x 0.053 +0.56 x 0.0537 + 3.28 x 0.0468

+ (0.0219-0.0)/12 x 100

= 0.1685 + 0.1921 + 0.1825

= 0.5431 kg

Sulphur balance error:

$$= \frac{\text{sulphur in-sulphur out}}{\text{sulphur in}} \times 100\%$$

$$= \frac{0.5575-0.5431}{0.5575} \times 100\%$$

= 2.58%

Sulphur capture

$$= \frac{\text{sulphur in fuel - sulphur in gas}}{\text{sulphur in fuel}} \times 100\%$$

= 69.8%

Calcium in:

= mass flow of limestone x mass fraction calcium

= 3.73 x 0.3734

= 1.3928 kg

Calcium out:

= $\sum$ mass flow of solids discharged x mass fraction of calcium in the solids
+ accumulation of calcium in the standpipe bed

= 0.16 x 0.2002 + 0.56 x 0.1025 + 3.28 x 0.1970
+ (0.0816 - 0.0)/12 x 100

= 0.7356 + 0.680
= 1.4156 kg

Calcium balance error:

$$= \frac{\text{Calcium in - Calcium out}}{\text{Calcium in}} \times 100\%$$

$$= \frac{1.3928 - 1.4156}{1.3928} \times 100\%$$

= -1.64%

Ash in:

= mass flow of coal x mass fraction ash
+ mass flow of limestone x (mass fraction ash + mass fraction CaO)
+ molar flow of sulphur captured by the discharged solids x molecular weight of SO_3

= 9.53 x 0.1502 + 3.73 x (0.0676 + 0.522)
+ (0.5575 - 0.1685)/32.06 x 80.06

= 1.431 + 2.199 + 0.971
= 4.601 kg

Ash out:

= $\sum$ mass flow of discharged solids x mass fraction ash in the solids

= 0.16 x 0.824 + 0.56 x 0.759 + 3.28 x 0.771

$$= 3.086 \text{ kg}$$

Ash balance error:

$$= \frac{\text{Ash in - Ash out}}{\text{Ash in}} \times 100\%$$

$$= \frac{4.601 - 3.086}{4.601} \times 100\%$$

$$= 32.9\%$$

Combustion efficiency:

$$= \frac{\text{Carbon in - carbon out in the solids}}{\text{carbon in}} \times 100\%$$

$$= \frac{6.9955 - 0.790}{6.995} \times 100\%$$

$$= 88.7\%$$

Air requried for complete combustion:

$$= \frac{\frac{32.00}{12.01} \times .6902 + \frac{8.00}{1.01} \times .0484 + \frac{32.00}{32.06} \times .0585 - 0.0337}{0.233}$$

$$= 9.644 \text{ kg air/kg coal}$$

Excess air:

$$= \frac{\text{air supplied}}{\text{air requried for complete combustion}}$$

$$= \frac{96.33/9.53}{9.644}$$

$$= 1.048$$

APPENDIX C

Calculation of Steady State Bed Composition

Calculation of Steady State Bed Composition

Feed: 100 kg coal + 39.2 kg elmtree limestone Ca/S ratio of 2.0

Assumptions: 100% combustion ie no carbon in the bed
85% sulphur capture

Products of combustion:

1) 100 x 0.1502 = 15.02 kg coal ash
2) 39.2 x (1-.9324) = 2.65 kg limestone ash
3) .085/2 x 39.2 x 0.3734 x 136.14/40.08 = 21.13 kg $CaSO_4$
4) 1.15/2 x 39.2 x 0.3734 x 56.08/40.08 = 11.78 kg CaO

Total products: = 15.02 + 2.65 + 21.13 + 11.78 = 50.58 kg

therefore Bed composition = 34.9% ash
41.8 % $CaSO_4$
23.3% CaO

the SiO_2 content in the steady state bed

= (15.02 x .30)/50.58
= 8.9%